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Importance of policy support and feedstock prices on economic feasibility of bioethanol production from wheat straw in the UK

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ABSTRACT

The economic feasibility of producing bioethanol from wheat straw in the UK using various state-of-the-art pretreatment technologies (steam explosion with and without acid catalyst, liquid hot water, dilute acid and wet oxidation) is assessed in this study. Under the current-technology base-case modeled using high enzyme loadings demonstrated at the laboratory-scale, wet oxidation pretreatment had the lowest minimum ethanol selling price (MESP) of £0.347/L ($2.032/gal). A contribution analysis showed feedstock price and enzyme cost were the two greatest contributors to the MESP, which led to a prospective case study and sensitivity analysis for assessing the effects of these two factors on the potential for economically competitive wheat straw-to-bioethanol UK supply chains. Prospective case studies modeled with a reduced enzyme loading and cost, demonstrated that although pretreatment scenarios with liquid hot water and steam explosion without acid catalyst were the closest to petrol pump prices, policy support in the form of tax exemptions could significantly enhance competitiveness of bioethanol with conventional fuel. A sensitivity analysis of feedstock prices also demonstrated that access to wheat straw prices of £35/t or lower would allow bioethanol production to be competitive with petrol under the best case scenario.

Contents

Abbreviations: CHP, Combined heat and power; COD, Chemical oxygen demand; CSL, Corn steep liquor; DA, Dilute acid; DA, Diammonium phosphate; FPU, Filter paper unit; GHG, Greenhouse gas; HMF, 5-hydroxymethyl-furfural; ISBL, Inside-battery-limits; LHW, Liquid hot water; MESP, Minimum ethanol selling price; NREL, National renewable energy laboratory; SE, Steam explosion; SECA, Steam explosion with sulphuric acid catalyst; TCI, Total capital investment; WO, Wet oxidation; WWT, Wastewater treatment; VAT, Value-added tax

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1. Introduction

The transportation sector is under increasing pressure to improve vehicle efficiency as well as to diversify transport fuel sources as approaches to reduce reliance on fossil fuels. Alternative transport fuels such as natural gas, hydrogen and biofuels are seen potential routes to reduce energy insecurity and environmental pressure. However, the production of biofuels from lignocellulosic biomass is considered to be the most promising option in the short-term as their market maturity is above the other alternatives [1,2]. In 2010, transport accounted for 21% of the UK's domestic GHG emissions—this was further increased to 26% when shipping and aviation were included [3,4]. The UK government's commitment to reducing transport-derived carbon emissions is driven by targets such as the EU Renewable Energy Directive which requires that 10% of transport fuel come from renewable sources by 2020, as well as the Renewable Transport Fuels Obligation which mandates that fuel suppliers increase the proportion of biofuels to reach 5% (by volume) of total road transport fuel by 2013 [4–6]. As a result of these policy measures, statistics reveal that domestic emissions, which rose steadily from years 1990 to 2007, have fallen back to their initial 1990 levels due to improvements in vehicle fuel efficiency and increased uptake of biofuels [4].

Agricultural wastes, as one type of lignocellulosic resource, can comprise up to 50% of agricultural production, and are regarded as cheap, abundant and accessible feedstocks for bioethanol production [7]. The UK produces around 11.9 million tonnes of cereal straw annually, with a 4.9 million tonne surplus [8]. Wheat straw accounts for 54% by mass of the total straw produced in the UK (oil seed rape 20%, barley 20%, oats 4% and others 1%) [9]. Wheat production in the UK ranks third amongst the EU 27 countries, with a total area covering 1.8 million hectares [10,11]. Though the composition of wheat straw compared to wood has been shown to have less cellulose and lignin and more hemicellulose, it still contains between 60% to 80% of its biomass composition as polysaccharides and thereby represents a significant potential resource of sugars for bioethanol production [9].

Numerous studies have investigated the technological feasibility of biochemically converting wheat straw into bioethanol, notably the effect that various pretreatments have on enhancing sugar conversion efficiencies [12–17]. A recent study by Talebnia et al. [17] has reviewed these technological routes and has suggested the commercial potential that bioethanol from wheat straw may have based on evaluations of techno-economic studies for other lignocellulosic feedstocks. From these reviews, it is generally concluded that pretreatment is a necessary step in the biochemical conversion process to ensure production of fermentable sugars within an industrially acceptable set of conditions [18,19]. Through combinations of increasing accessible surface area, partially/fully removing lignin and/or hemicellulose and disrupting interactions between cell wall components, an effective pretreatment aims to improve enzyme accessibility to glucan, thereby increasing the amount of sugars available for fermentation [20,21]. Successful pretreatments on wheat straw, of which include dilute acid, steam explosion, liquid hot water and wet oxidation amongst others, have been shown to enhance sugar yields after enzymatic saccharification, achieving up to 74% to 99.6% of the theoretical maximum [13,14,16,17].

Existing techno-economic assessments of bioethanol production from various feedstocks including corn stover, poplar, eucalyptus and waste papers amongst others have been widely reported in the literature; however there is yet to be an in-depth analysis of the economic viability of wheat straw for this purpose [22–31]. These assessments are useful tools for providing insight into possibilities for process optimisation, cost reduction, and comparison of alternative technology scenarios. In this work we have adapted and modified the techno-economic model in Humbird et al. [22] to build five pretreatment process scenarios for bioethanol production from wheat straw in the UK. The effects of these scenarios are evaluated using the minimum ethanol selling price (MESP) as the principal indicator of their economic feasibility.

2. Materials and methods

The composition of wheat straw, the conditions of the pretreatments applied on wheat straw and the results from the subsequent enzymatic saccharification are derived from research literature. This information has been used as input for the process design and simulation carried out using AspenPlus^{TM} software. The generated mass and energy balances are used for further economic analysis to assess the economic feasibility of bioethanol produced from wheat straw for comparison with conventional transportation fuel petrol.

2.1. Composition of wheat straw

The compositional data of wheat straw derived from research literature has been normalised and averaged in this study [13,14]. Wheat straw, with a moisture content of 6.5% (w/w), contains 34.6% glucan, 21.1% xylan, 2.3% arabinan, 0.9% galactan, 18.0% lignin, 2.2% acetyl groups, 5.6% ash and 15.4% extractives (w/w on a dry basis).

2.2. Pretreatment and saccharification

A recent comprehensive review by Talebnia et al. [17] summarised the pretreatment methods on wheat straw which have been modelled in this study. The conditions of these pretreatments and results for the subsequent enzymatic saccharification are listed in Table 1. It should be noted that in order to complete the process design, some assumptions have been made in the present work for the information that was not reported in the literature. For example, it is assumed that there is no 5-hydroxymethyl-furfural (HMF) formed in liquid hot water and wet oxidation pretreatments, and no xylan is degraded into furfural in liquid hot water pretreatment. When not reported, the conversion yields of arabinan and galactan during pretreatment have also been assumed to be the same as xylan and glucan, respectively. Similarly, C5 and C6 sugar yields in enzymatic saccharification have also been assumed to be the same as xylose and glucose yields, respectively. Where only the glucose yield is reported, it is assumed that other polysaccharides have the same conversion efficiencies.

Table 1
A summary of conditions and results for the selected pretreatment and enzymatic saccharification processes.

Pretreatment method	Liquid hot water (LHW)	Steam explosion without catalyst (SE)	Steam explosion with H_2SO_4 catalyst (SECA)	Dilute acid (DA)	Wet oxidation (WO)
Pretreatment conditions	188 °C, 40 min	180 °C, 10 min	180 °C, 10 min, 0.9% (w/w) H_2SO_4	121 °C, 90 min, 2% (w/v) H_2SO_4	180 °C, 15 min, O_2 at 12 bar
Pretreatment reactions	Fraction of reactant converted to product				
Glucan + H_2O → Glucose	0%	10%	10%	12%	1.2%
Glucan → HMF + H_2O	NA [a]	1.5%	0.15%	2.7%	NA
Xylan + H_2O → Xylose	91%	95%	82%	85%	7.1%
Xylan → Furfural + $2H_2O$	NA	0.35%	0.05%	0.2%	1.2%
Arabinan + H_2O → Arabinose	NA	94%	90%	100%	NA
Galactan + H_2O → Galactose	NA	100%	100%	NA	NA
Lignin → Soluble lignin	19%	18%	32%	16%	NA
Enzymatic saccharification conditions	(15 FPU cellulase + 15 IU β-glucosidase)/g dry solid in treated biomass, 72 h	(15 FPU cellulase + 12.6 IU β-glucosidase)/g dry solid in treated biomass, 72 h	(15 FPU cellulase + 12.6 IU β-glucosidase)/g dry solid in treated biomass, 72 h	(20 FPU cellulase + 40 IU β-glucosidase)/g glucan, 72 h	10 FPU/g glucan of enzyme mixture (3:1 v/v ratio of cellulase: β-glucosidase), 72 h
Enzymatic saccharification results	76.3% glucose yield, 20.5% xylose yield [b]	60% glucose yield	80% glucose yield	62.2% glucose yield, 73.9% xylose yield	56% glucose yield, 65% xylose yield

[a] NA = Data is not reported. Assumptions are applied in process designs (see Section 2.2).
[b] Sugar yield is defined as the proportion of polysaccharides converted to monomeric sugars. Assumptions are made when applicable (see Section 2.2).

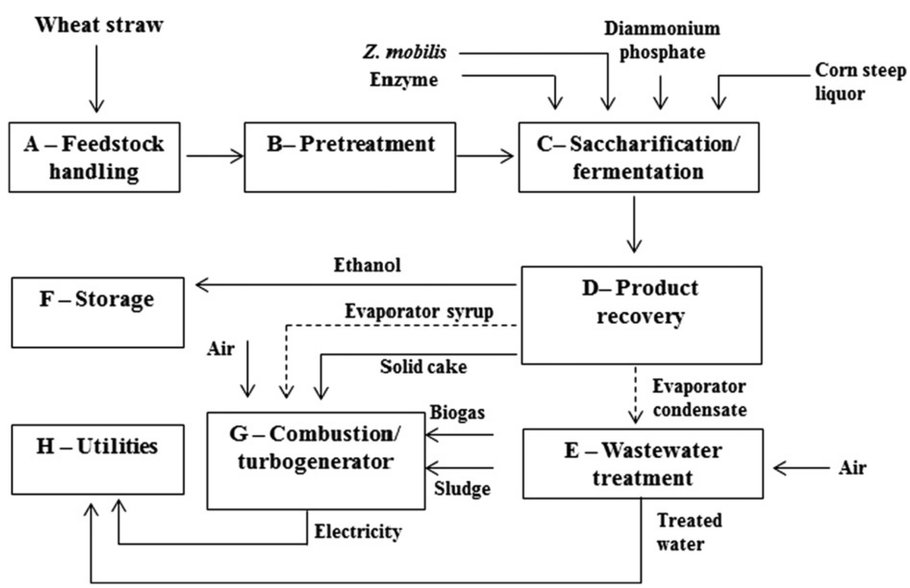


Fig. 1. A schematic diagram of the wheat straw-to-bioethanol process (streams shown in dashed lines vary in scenarios depending on process design differences in Area D).

2.3. Process design and simulation

The process design configurations have been developed based on the NREL corn stover-to-bioethanol process utilising dilute sulphuric acid pretreatment [22]. The plant size is designed to handle 2000 dry metric tonnes of wheat straw per day [22]. An overview of the wheat straw-to-bioethanol process is described in the schematic diagram in Fig. 1.

The process starts from the feedstock handling area (Area A) where wheat straw is unloaded, unwrapped, washed and milled to a suitable particle size and conveyed to the pretreatment section (Area B). The total solid loading of pretreatment is designed as 30% (w/w) [22]. Detailed designs in Area B based on the selected pretreatment methods are described in Section 2.3.1. The pretreated wheat straw is sent to the separate saccharification and fermentation area (Area C) where polysaccharides are hydrolysed by enzymes (cellulase and β-glucosidase mixture) to monomeric sugars and then fermented into ethanol by the fermentative bacterium, *Zymomonas mobilis*. Saccharification is carried out at 50 °C for 72 h to give the monomeric sugar yields listed in Table 1. The hydrolysate is cooled to 32 °C and sent to (1) two *Z. mobilis* seed inoculation trains with a residence time of 24 h each and (2) fermentation tanks operating for 36 h. *Z. mobilis* used for fermentation in our design, is a recombinant micro-organism which ferments both hexose and pentose sugars. Nutrients including corn steep liquor (CSL) and diammonium

phosphate (DAP) are required in both seed inoculation and fermentation. Nutrient loadings and fermentation sugar conversion efficiencies (95% for glucose, 85% for xylose and arabinose) are adopted from the NREL process [22]. Of the total monomeric sugars, 3% are assumed to be converted into glycerol, succinic acid and xylitol due to contaminations [22].

The data provided in the table are derived from the following references: Pérez et al. [13], Ballesteros et al. [14], Qi et al. [15] and Georgieva et al. [16].

The fermentation beer is fed to the product recovery area (Area D) where ethanol is concentrated through distillation and molecular sieve adsorption to 99.6%. The distillation bottoms obtained from the distillation column contain unfermented monomeric sugars, organic acids and solid residuals such as lignin, extractives and ash. These distillation bottoms are either sent to: (1) a series of evaporators to concentrate the soluble organics as a syrup which is then fed to the combustor (Area G) for heat generation; or (2) a press filter to separate solids which are sent to the combustor while the liquor is sent to wastewater treatment (Area E). The different pretreatment processes have major impacts on the design of downstream processes and these are described in Section 2.3.2.

The wastewater treatment area (WWT) (Area E) includes anaerobic and aerobic digestion which treats used water and recycles it within the process to minimise the amount of water discharged to the environment as well as the purchased fresh water requirement. In anaerobic digestion, 91% of organic matter is converted into biogas and microorganism cell mass. The biogas with a composition of 51% CH_4/49% CO_2 (dry molar basis) is produced at a yield of 228 g biogas/kg COD (chemical oxygen demand) removed. The treated water is further cleaned in aerobic digestion where 96% of the remaining soluble organic matter is removed. Wastewater treatment chemicals such as caustic soda are involved in the processes where ammonia is used in pretreatment for acid neutralisation i.e., steam explosion with catalyst and dilute acid pretreatments [22]. In these cases, the cleaned water is then sent to a reverse osmosis membrane system for salt removal (sodium nitrate). The sodium nitrate is modelled as brine waste sent to landfill [22].

The concentrated syrup where applicable, combined with the solid cake separated from the distillation bottoms, biogas and cell mass (sludge) obtained from wastewater treatment are fed to the combustor (Area G) for combined heat and power (CHP) generation. Boiler and generator efficiencies are assumed to be 80% and 85%, respectively. High pressure steam is extracted from the turbine to meet process heat requirements. The generated electricity is used to supply the process, and any excess can be sold to the National Grid as a co-product credit. For the pretreatments involving acid (SECA and DA), the flue gas released from the combustor requires desulphurisation by applying limestone before emitting to the atmosphere.

The utilities area (Area H) is responsible for the cooling tower system, clean-in-place system and plant air system. Feedstock, chemicals, and products are stored in the storage area (Area F).

2.3.1. Detailed process designs for pretreatment area (Area B)

2.3.1.1. Steam explosion without catalyst (SE). Wheat straw from the feedstock handling area (Area A) is conveyed to Area B, mixed with process water recycled from WWT (Area E), and heated up to 100 °C by high pressure steam (13 atm). The flow is then sent to the pretreatment reactor where high pressure steam is injected to maintain the reactor temperature at 180 °C. The pretreatment is carried out for 10 min before a sudden pressure release followed by sending the flow to a blowdown tank to vaporize acetic acid and other inhibitory organics generated during pretreatment. The flash vapour is then condensed via heat exchange, with the stream going to the distillation column and then to WWT.

2.3.1.2. Steam explosion with H_2SO_4 catalyst (SECA). Similarly to the steam explosion without catalyst pretreatment, the wheat straw is pre-heated and sent the pretreatment reactor where a stream of diluted sulphuric acid is injected to achieve a concentration of 0.9% (w/w). After the blowdown tank, the flow is sent to a conditioning tank where ammonia is added to neutralise the hydrolysate to a suitable pH for saccharification.

2.3.1.3. Dilute acid (DA). The process design is similar to that of steam explosion with catalyst. The pretreatment reactor operates at 121 °C by adding high pressure steam. Blowdown and conditioning tanks are also required before sending the detoxified hydrolysate to the saccharification tank.

2.3.1.4. Liquid hot water (LHW). Wheat straw is diluted with water and pre-heated to 100 °C before holding for 40 min at 188 °C in the pretreatment reactor. The slurry is then transferred directly to the saccharification tank.

2.3.1.5. Wet oxidation (WO). After wheat straw is mixed with process water and pre-heated to 100 °C, O_2 is compressed to 12 bar and injected into the pretreatment reactor which operates at 180 °C for 15 min. In this process design, blowdown and conditioning tanks are excluded.

The process designs are grouped into three scenarios shown in Fig. 2: (a) process design for LHW and SE, (b) process design for SECA and DA, and (c) process design for WO.

2.3.2. Process design for product recovery area (Area D)

2.3.2.1. Design including evaporators (Fig. 3(a)). This design applies to LHW, SE, and WO pretreatment scenarios. The distillation bottoms are sent to the first evaporator where approximately 30% of water in the stream is evaporated by heat provided by high pressure steam. The stream is then sent to a press filter to separate the lignin-rich solid from the liquid fraction. The liquid fraction is fed to the second evaporator where a further 45% (approx.) is evaporated and then to the third evaporator where a condensed syrup is obtained after approx. 80% of the remaining water is removed. The solid cake is combined with the syrup and this stream which contains a moisture content of less than 45% is sent to the combustor [22].

2.3.2.2. Design excluding evaporators (Fig. 3(b)). In SECA and DA pretreatment scenarios, the distillation bottoms are sent to the press filter directly for solid–liquid separation. The solid cake is conveyed to the combustor while liquid is sent to WWT.

2.4. Economic assessment

2.4.1. Cost estimation

Once mass and energy balances for the process have been generated, the economics of bioethanol production are determined using a standard cost estimation method adopted from NREL [22]. This cost estimation method is widely used in techno-economic assessments of biofuel production from lignocellulosic feedstocks utilising various processing technologies [25,27,29,31,32]. The total capital investment (TCI) is determined from equipment purchased and installation costs estimated from process specifications. Equipment costs derived from NREL's vendor quotations reflect a baseline equipment size which can be scaled up or down according to the

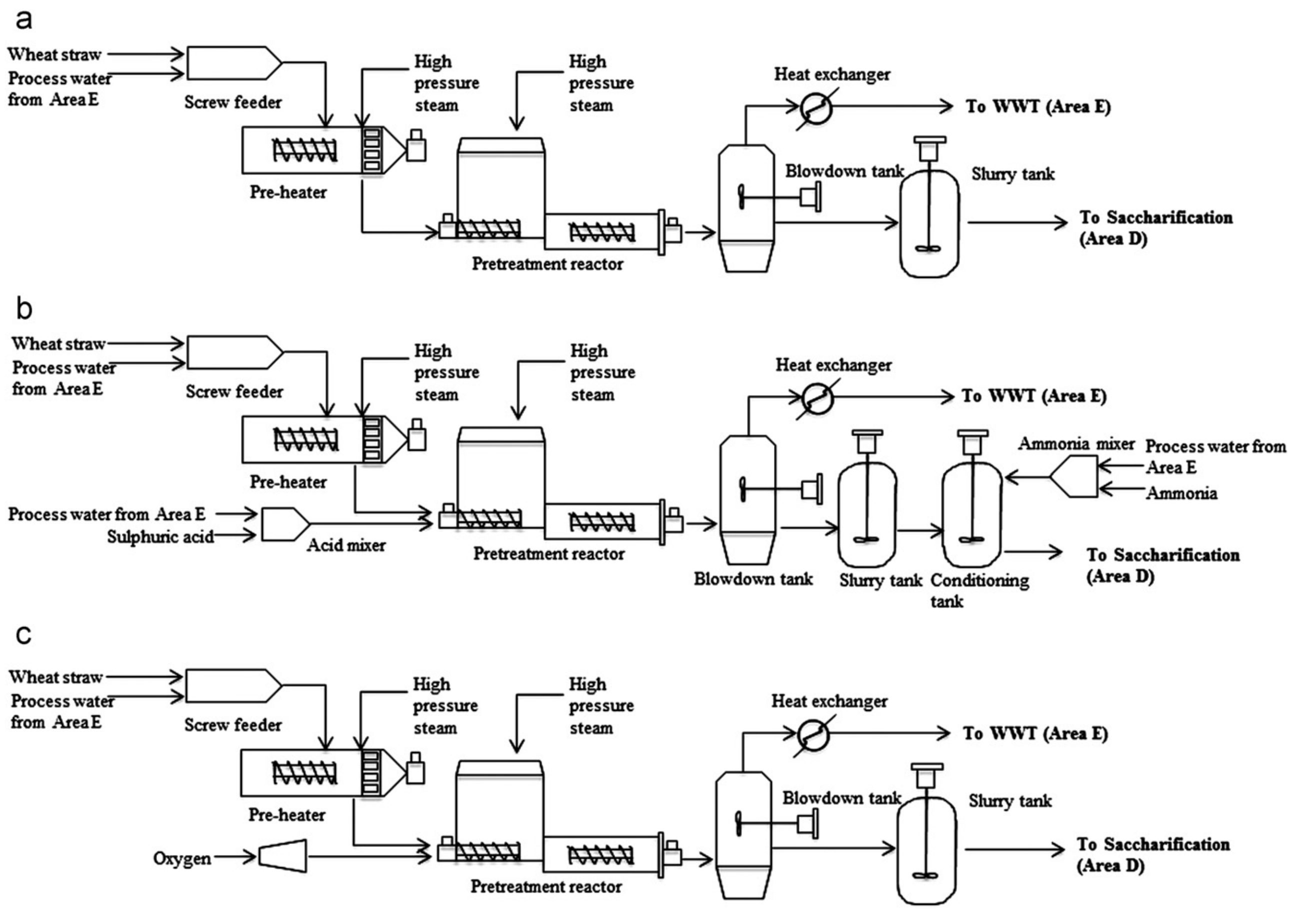


Fig. 2. Process designs for different pretreatment methods in Area B ((a) for SE and LHW (b) for DA and SECA; (c) for WO).

Table 2
Summary of variable operating costs.

Materials/chemicals/energy	Cost/price	Reference
Wheat straw	45.7 £/tonne [a]	[33,34]
Sulphuric acid	55.8 £/tonne	[35]
Ammonia	295.5 £/tonne	[22]
Lime ($Ca(OH)_2$)	131.3 £/tonne	[22]
Oxygen (O_2)	64.6 £/tonne	[36]
Corn steep liquor	37.4 £/tonne	[22]
Diammonium phosphate (DAP)	323.6 £/tonne	[37]
Enzyme (Celluclast 1.5L+Novozyme 188)	327.7 £/tonne	[27]
Sorbitol	742.0 £/tonne	[22]
Caustic	98.5 £/tonne	[38]
Fresh water	0.17 £/tonne	[22]
Boiler feed water chemicals	2758.8 £/tonne	[22]
Cooling tower chemicals	2350.6 £/tonne	[22]
Ash landfill	24.5 £/tonne[b]	[29,39]
Surplus electricity	0.093 £/kW h	[40]

[a] Includes a transportation cost of £7.7 up to distances of 113 km.
[b] Includes landfill cost of £22/tonne and tax £2.5/tonne.

exponential scaling expression [22]:

$$\text{New cost} = (\text{Base cost})\left(\frac{\text{New size}}{\text{Base size}}\right)^{f\ \text{scale}}$$

The equipment cost is then indexed to the reference year of 2010 chosen for this study. Once the total equipment costs have been calculated in the year of interest, direct and indirect costs are added to yield the TCI. Direct costs include warehouse, site development and additional piping, which comprise 4%, 9% and 4.5% of the inside-battery-limits (ISBL) equipment costs (Areas B–D), respectively. Indirect costs include proratable costs (10% of total direct cost), field expenses (10%), home office and construction (20%), project contingency (10%) and other costs (10%) [22].

The variable operating costs including raw materials, waste handling charges and by-product credits, are only incurred while the process is operating (Table 2). Fixed operating costs include

Table 3
Discounted cash flow parameters.

Parameters	Value
Plant life	30 years
Discount rate	10%
Financing	40% equity
Loan terms	10-year loan at 8% APR
General plant depreciation	200% declining balance [a]
General plant recovery period	7 years
Steam plant depreciation	150% declining balance
Steam plant recovery period	20 years
Corporation tax rates	28% [b]
Construction period	3 years
0–12 months	8% of project cost
12–24 months	60% of project cost
24–36 months	32% of project cost
Working capital	5% of fixed capital investment
Start-up time	3 months
Revenues during start-up	50%
Variable costs incurred during start-up	75%
Fixed costs incurred during start-up	100%

[a] Depreciation method is the IRS modified accelerated cost recovery system (MACRS) [22].
[b] From HM Revenue & Customs UK [41].

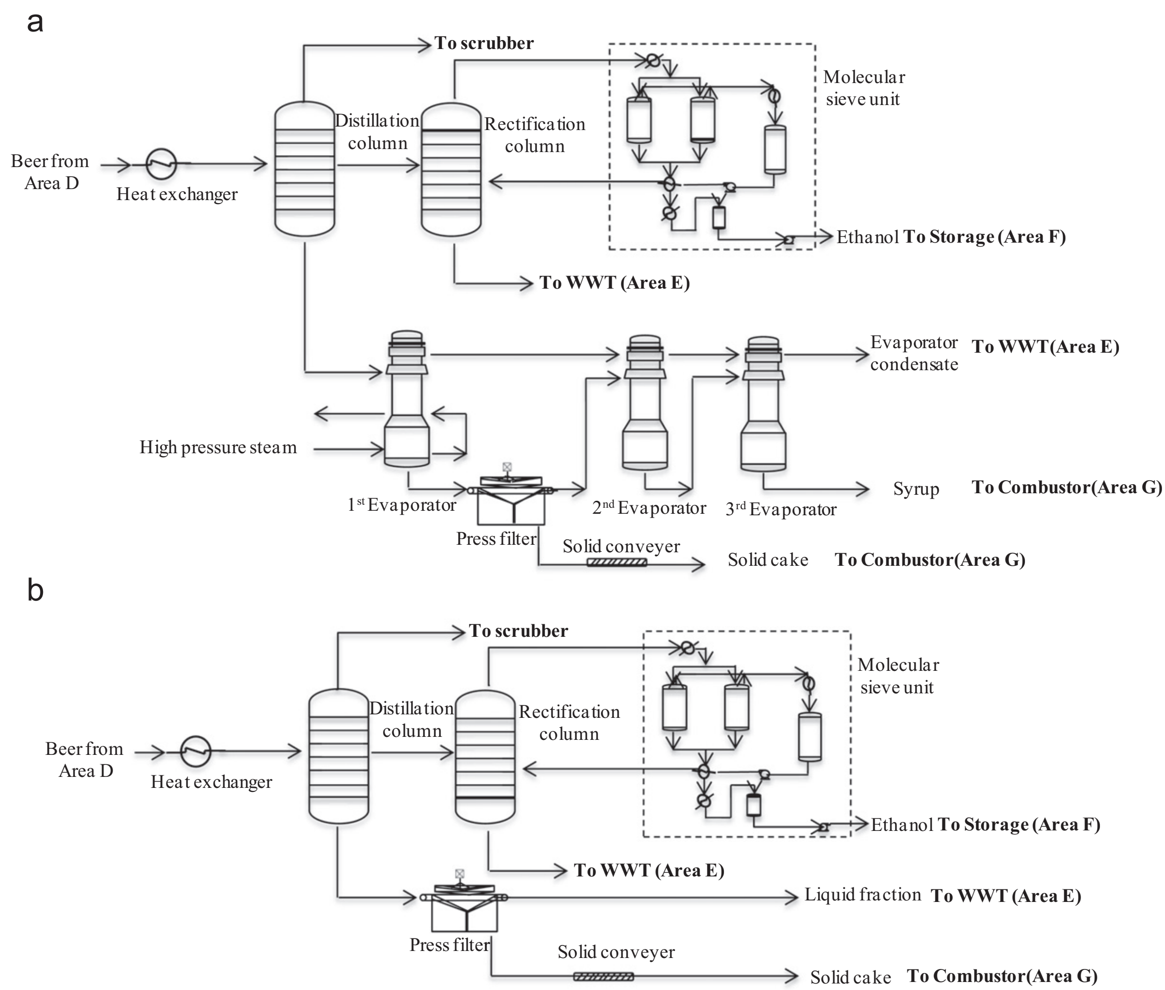


Fig. 3. Process designs for alternative outcomes for the distillation bottoms in Area D ((a) Design including evaporators and (b) design excluding evaporators).

labour and various overhead items—these are incurred whether or not the plant is producing at full capacity. The number of employees and salaries are derived from a UK-based case study [22,33]. Annual maintenance materials are estimated as 3% of the ISBL capital cost, local property tax and property insurance are assumed to be 0.7% of the fixed capital investment [22].

2.4.2. *Discounted cash flow method*

Once the total capital investment and operating costs have been determined, a discounted cash flow method is used to estimate the minimum ethanol selling price (MESP). The MESP refers the bioethanol price at which the net present value of the project is zero at a set discount rate of 10%. This model is based on 'nth-plant' assumption which assumes several plants using the same technology are currently operating—this eliminates additional costs associated with pioneer plants [22]. The parameters used in the discounted cash flow calculation are listed in Table 3.

2.5. *Supply-chain model*

In order to compare wheat straw-derived bioethanol with petrol, the supply-chain model is established to determine the bioethanol price at pump. This price includes the bioethanol cost of production, fuel duty (58.19 p/L in 2010 [42]), value-added tax (VAT) (17.5% in 2010), a feedstock transportation cost and a fuel distribution cost of £0.021/L. It is assumed that wheat straw is transported by lorry from a distance within a 50 km radius of the bioethanol plant. An average transportation and handling charge of £7.7 is adopted from Hess et al. [34] for distances up to 113 km (Fig. 3).

Table 4
Summary of economic results for pretreatment scenarios.

Pretreatment	MESP (£/L)	Ethanol production (MM L/yr)	Ethanol yield (L/MT)	Electricity generated (MW)	Electricity credit (MM£/yr)
WO	0.347	182	237	48.0	20.7
SECA	0.488	223	290	24.7	1.2
SE	0.453	196	255	46.0	20.2
LHW	0.474	208	270	39.4	15.4
DA	0.486	201	261	37.0	10.6

3. Results and discussion

3.1. *Comparison of the MESP and total cost breakdown between five scenarios*

A summary of economic results are presented in Table 4. The process with the lowest MESP is WO, which is £0.347/L ($2.032/gal), with the others ranging from £0.453/L to £0.488/L ($2.653/gal to $2.858/gal). Ethanol yield varies between 237 L/MT and 290 L/MT in the various processes. Ethanol production and yield is highest under the

SECA process, which also has the lowest electricity generation and credit.

The cost breakdown for each process area and MESPs of bioethanol derived from wheat straw under five pretreatment scenarios are shown in Fig. 4. Although the total cost of SECA is the second lowest amongst the different pretreatments, the amount of electricity credit gained is also the least, therefore offsetting the overall cost by very little, resulting in this process having the highest net MESP. The two greatest cost contributors amongst the process scenarios are the feedstock price (included in feedstock handling area) and the enzyme price (included in saccharification/fermentation area). Feedstock price contribution varies between the scenarios, but accounts for approximately 36–56% of the MESP. Although the price of wheat straw has been as low as £30/tonne in 2009, a 20–40% increase in price occurred in 2010 and the straw price is also highly subject to regional variations around the UK [43]. The effects of MESP and bioethanol price at pump caused by fluctuation of wheat straw prices are discussed in Section 3.3.

Enzyme price contributes 18–43% of the MESP depending on the enzyme loading defined in enzymatic saccharification. This is shown in the Saccharification/fermentation contribution in Fig. 4, which is relatively larger for the scenarios with higher enzyme loadings (SE, LHW and DA) in comparison with WO and SECA. In the highest three scenarios, the enzyme contribution is even greater than feedstock contribution, highlighting the importance of minimising enzyme loading during saccharification in order to reduce the overall MESP. Although enzyme producers have undertaken significant efforts to reduce enzyme cost, it is evident that this parameter continues to be a major factor in the commercialisation of cellulosic ethanol [44]. By including the potential improvements in enzyme production (in terms of higher activity and therefore lower loading or reduced price), a prospective case study is performed in Section 3.3.

Besides feedstock and enzyme costs, pretreatment and WWT are also recognised to make relatively significant contributions to MESPs. Therefore further insights on these two processes are performed in Sections 3.2 and 3.3.

3.2. *Pretreatment area cost breakdown and effect on downstream processes*

Particular attention has been drawn to the pretreatment stage due to three main reasons. First, this stage is highly variable and subject to different process designs and conditions, therefore it is necessary to understand which technologies are more viable from an economic perspective. Second, by implementing an effective pretreatment which can improve the digestibility of the biomass to enzymatic saccharification, this can potentially decrease enzyme loading and hence cost of saccharification which is a major contributor to the MESP. And third, pretreatment selection may have a large impact on the design of downstream processes. A cost breakdown demonstrates the relative contributions of capital, chemicals and waste, electricity and fixed costs between pretreatment scenarios (Fig. 5).

DA pretreatment has the greatest total cost of £0.09/L ($0.528/gal) followed by WO at £0.08/L ($0.468/gal). The lowest costing pretreatments are LHW and SE at £0.046/L ($0.268/gal) and £0.049/L ($0.285/gal), respectively. From the minimal variation in capital recovery charge, process electricity and fixed costs between the five pretreatments, it is evident that the greatest cost contributor to the pretreatment stage is the cost of chemicals and waste.

In DA and SECA scenarios, purchase of sulphuric acid as well as the subsequent hydrolysate neutralisation with ammonia gas accounts for 56% and 39% of the pretreatment area, respectively. It is also important to note the significant impact that acid concentration has on the cost, shown by the chemicals and waste contribution in DA at £0.05/L ($0.295/gal) which is double that of

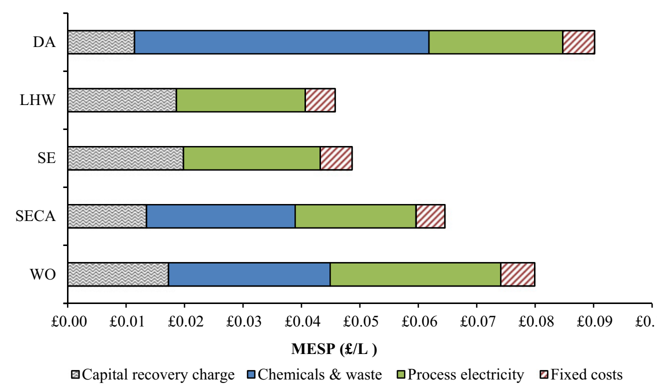


Fig. 5. Cost breakdown of the pretreatment stage for five process scenarios.

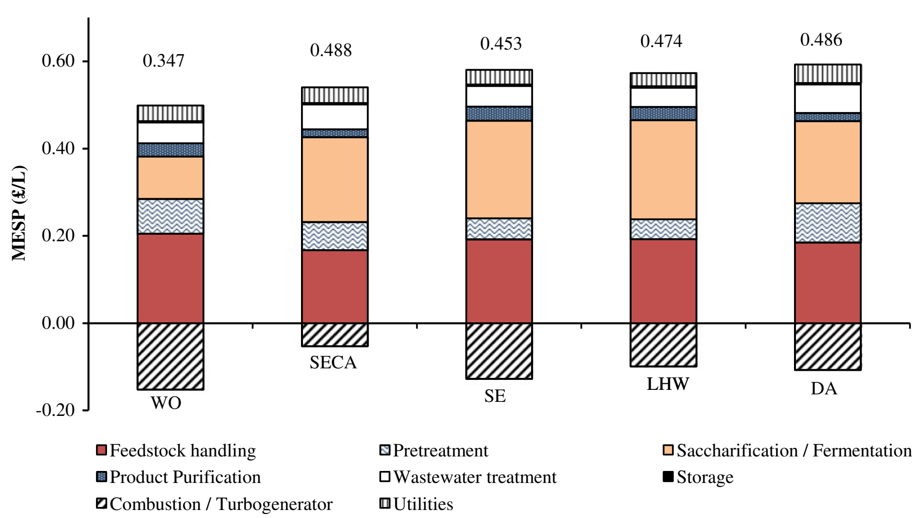


Fig. 4. Summary of net MESPs and total process cost breakdown for five pretreatment scenarios.
Note: WO—wet oxidation, SE—steam explosion, SECA—steam explosion with acid catalyst, DA—dilute acid, LHW—liquid hot water. Net MESP values indicated above each scenario bar.

SECA at £0.025/L ($0.149/gal) when a more concentrated acid is used (2% w/v compared to 0.9% w/w). WO has the second highest cost, primarily due to the price of oxygen purchase (included in chemicals and waste), but also due to greater process electricity consumption (included in process electricity) from the oxygen compressor required to reach a pressure of 12 bar.

The effects of different pretreatments on downstream costs, as well as water usage, heat usage and electricity generation are noted. As stated in Section 2.3, ammonia neutralisation of pretreatment hydrolysate in DA and SECA scenarios result in the formation of ammonium acetate and sulphate salts which, when concentrated using evaporators, result in combustor fouling. The replacement of the evaporator system with a solid/liquid separation system leads to a reduced product recovery area cost from approximately £0.031/L to £0.019/L ($0.180/gal to $1.110/gal) [22]. Furthermore, the additional salt removal system in wastewater treatment raises capital costs within this area from approximately £0.046/L to £0.061/L ($0.272/gal to $0.363/gal), demonstrating the impact that pretreatment can have on other areas within the conversion process.

Electricity generation in the different scenarios in order of lowest to highest are: SECA (25 MW), DA (37 MW), LHW (39 MW), SE (46 MW) and WO (48 MW), as shown in Table 4. Electricity generation is correlated with saccharification yields, shown clearly by the highest conversions (80% of glucan) in SECA which produces the least amount of electricity. With the exception of SECA which only has 1.6 MW of surplus electricity, all the other scenarios are able to sell 37–55% of their total electricity generation to the grid. Greatest heat requirements came from scenarios employing the evaporator system (WO, SE, LHW), resulting in higher process water requirements over those with solid/liquid separation (SECA, DA).

3.3. *Comparison of the prospective case MESP with petrol and barriers for commercialisation*

The cost breakdown in Section 3.1 clearly indicates the major contribution that enzyme cost and feedstock price have on the MESP. Prospective pretreatment scenarios with reduced enzyme loadings and costs are therefore modelled to evaluate the potential for reducing the MESP with assumed technological advances expected to take place, as well as to allow a more direct comparison between pretreatment scenarios under consistent conditions. The prospective case assumes a 75% reduction in enzyme cost [45], as well as a lowered enzyme loading to 10 FPU/g glucan.

In the prospective cases, SE has the lowest MESP followed by LHW, WO, SECA and lastly DA. There is a significant MESP reduction in the prospective cases for LHW and SE from £0.474/L to £0.281/L ($2.776/gal to $1.646/gal) and £0.453/L to £0.278/L ($2.653/gal to $1.628/gal), respectively, whereas a much smaller decrease occurs in the MESP of WO from £0.347/L to £0.299/L ($2.032/gal to $1.751/gal). The greater reduction seen in LHW and SE is due to their high enzyme loading in the base-case scenario, which is significantly reduced in the prospective case. WO uses 10 FPU/g glucan in the base-case which means the decreased MESP is attributed solely to the reduction in enzyme cost.

To compare the cost competitiveness of bioethanol derived from wheat straw under different pretreatment scenarios, the current and prospective scenarios are compared against a reference petroleum price of premium unleaded petrol from the UK in 2010. The minimal selling price of bioethanol at the pump comprises the MESP, a fuel tax (includes fuel duty and 17.5% VAT) as well as a fuel distribution cost. The energy content of bioethanol is less than petrol (21.2 MJ/L compared to 31.2 MJ/L) such that 1 L of bioethanol is equivalent to 0.68 L of petrol. The bioethanol prices at pump have therefore been corrected to the equivalent price of petrol. Fig. 6 shows the cost breakdown of the pump price of bioethanol for current and prospective cases under the five pretreatment scenarios.

None of the wheat straw bioethanol baseline scenarios are competitive with the average petrol pump price of £1.17/L in 2010. In the prospective cases, SE and LHW are the closest to the petrol price at £1.18/L and £1.19/L, respectively. Although the production cost of bioethanol from wheat straw is fairly low, when a distribution cost (£0.021/L), fuel duty (58.19 p/L) as well as VAT (17.5%) are factored in, the MESP of bioethanol at the pump is more than double the cost of its production.

The retail petrol price in the UK is amongst the top five in the European Union, and also levies the highest fuel duty [46]. Fiscal incentives once offered biofuel production a selective advantage in comparison with petrol via a 20 p/L duty derogation. Even now, this scheme, which was described as "pale in comparison" to Germany's 100% tax exemption, was dropped in April 2010 in support of an equivalent fuel duty to petrol of 58.19 p/L [47]. Furthermore, the reversion of VAT rates from 15% back to 17.5% imposed on the fuel cost plus duty, only dampens competitiveness of bioethanol in relation to conventional fuels. If the 20 p/L differential rate for bioethanol was reinstated, all of the prospective cases ranging from £0.953/L to £1.059/L would be competitive with petrol, and WO would be competitive under

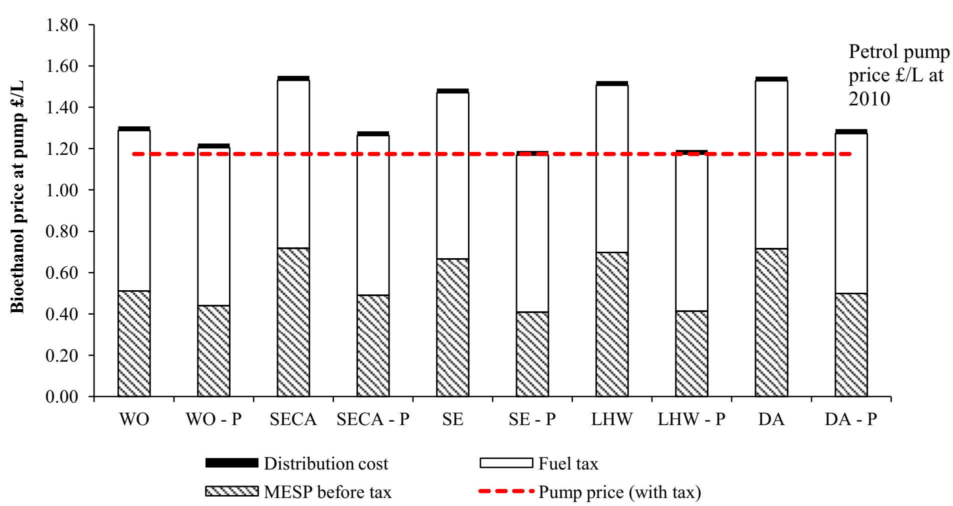


Fig. 6. Comparison of bioethanol price at pump derived from wheat straw with petrol under five current and prospective process scenarios (marked 'P').

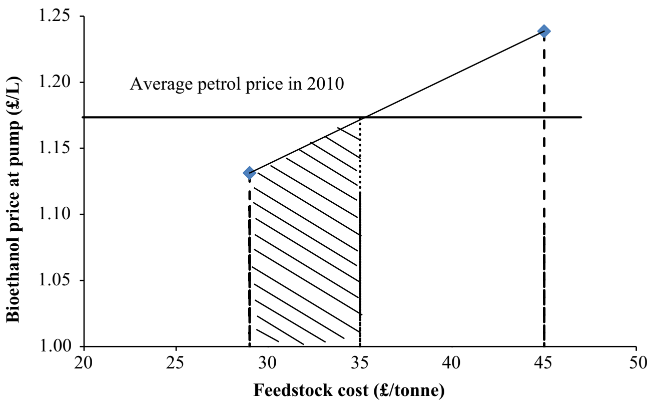


Fig. 7. Sensitivity analysis of bioethanol pump price produced under SE pretreatment to the lower and upper limits of wheat straw feedstock cost reported in 2010.

the base-case scenario at £1.073/L. These figures are clear indicators for policy-makers that a guarantee for future tax exemptions and obligations are likely to be needed to overcome the challenge of margin management as well as to allow greater market certainty for development of a wheat straw-based biofuel industry in the UK [47].

As noted in Section 3.1, feedstock price is one of the biggest costs, accounting for 37–56% of the total MESP. In 2010 the price of wheat straw fluctuated between £29/tonne and £45/tonne (excluding transportation cost of approx. £7.7/tonne) [22]. In order to understand the sensitivity of the bioethanol pump price to feedstock cost, we have applied these lowest and highest feedstock prices under the best modelled prospective case (SE pretreatment) (Fig. 7). At a price of £38/tonne used in the current analysis, the MESP is £1.18/L which is slightly higher than the average petrol price in 2010 of £1.17/L. The analysis shows that at a cost of £35/tonne and lower, bioethanol production from wheat straw can compete with petrol. Therefore access to a lower feedstock cost offers a major advantage for producers to reduce the MESP of wheat straw-derived bioethanol.

The assumption that lignocellulosic feedstocks from agricultural and similar residues are merely cheap waste or by-products which have no other purpose other than to be disposed of is a misconception. Although there is clearly a surplus of straw in the UK, it serves an important purpose in numerous applications of which maintaining soil condition by providing organic amendment to crops is one of the most significant [48]. Therefore it should not be assumed that farmers are seeking ways to dispose of their straw residues as it is valuable in enhancing soil productivity, as animal bedding and other uses. Consequently, the maximum amount of straw that can be diverted towards biofuel production will be largely determined by the balance of: (1) Retaining sufficient in-field to protect soil and water resources; (2) the margin that farmers can obtain after the additional costs of baling and removal; and (3) the cost of buying-in fertilizer for replacement of nutrients forgone in their removal [49]. With current farm-gate spot prices reported to be 84% of the value sold for biofuel production, the above factors may explain why the costs of agricultural wastes such as wheat straw may not be as cheap as expected. This is not to say that agricultural residues are not economically suited for biofuel production, but more so to suggest why wheat straw which is supposedly cheap, abundant and available in excess, nevertheless represents the single largest cost contributor to the cost of bioethanol. This is likely to remain the case in future years.

4. Conclusions

Techno-economic assessments of bioethanol derived from UK wheat straw with five pretreatment technologies were performed using data from literature, process simulation and expert consultation. In the current-technology base-case scenario modelled with high enzyme loadings, wet oxidation pretreatment had the lowest minimum ethanol selling price of £0.347/L. The contribution analysis shows that the main cost contributors across all scenarios were the wheat straw feedstock prices and the enzyme costs, ranging between 37–56% and 17–41%, respectively, of the MESP. Within the pretreatment area alone, the cost of chemicals was the most significant contributor, and as a result, liquid hot water and steam explosion without acid catalyst were the two scenarios with the lowest total area cost. It was also shown that pretreatments without chemical requirements (LHW, SE and WO) had a higher product recovery area cost due to the expensive evaporator system, but a lower wastewater treatment area cost attributed to the lack of a requirement for a salt removal system.

Prospective cases modelled with a lower enzyme loading and enzyme price, projected significant reductions in the MESP, particularly for the cases with liquid hot water or steam explosion without catalyst pretreatments. After factoring in taxes and distribution costs at 2010 prices, these two scenarios were the closest to being competitive with petrol (£1.17/L) at pump prices of £1.19/L and £1.18/L, respectively. The removal of fuel duty exemptions on bioethanol implemented in 2010 prevents all prospective cases from being economically competitive. A sensitivity analysis on wheat straw price showed that under the best prospective pretreatment (steam explosion without catalyst), access to wheat straw prices lower than £35/tonne would offer a significant advantage for producers to make bioethanol from wheat straw competitive (without duty derogation) with petrol. Taken as a whole, we conclude that while scientific research should still focus on ways to increase conversion efficiency (e.g., optimised pretreatments and reduced enzyme loadings), the overall supply-chain model indicates that there are greater obstacles such as wheat straw feedstock pricing and policy support that will be highly significant in enabling the development of a viable 2 G bioethanol industry based on UK wheat straw.

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